

Document made available under the Patent Cooperation Treaty (PCT)

International application number: PCT/US2006/009103

International filing date: 14 March 2006 (14.03.2006)

Document type: Certified copy of priority document

Document details: Country/Office: US
Number: 60/673,518
Filing date: 20 April 2005 (20.04.2005)

Date of receipt at the International Bureau: 03 May 2006 (03.05.2006)

Remark: Priority document submitted or transmitted to the International Bureau in compliance with Rule 17.1(a) or (b)



World Intellectual Property Organization (WIPO) - Geneva, Switzerland
Organisation Mondiale de la Propriété Intellectuelle (OMPI) - Genève, Suisse

1457988

THE UNITED STATES OF AMERICA

TO ALL TO WHOM THESE PRESENTS SHALL COME:

UNITED STATES DEPARTMENT OF COMMERCE

United States Patent and Trademark Office

April 25, 2006

THIS IS TO CERTIFY THAT ANNEXED HERETO IS A TRUE COPY FROM THE RECORDS OF THE UNITED STATES PATENT AND TRADEMARK OFFICE OF THOSE PAPERS OF THE BELOW IDENTIFIED PATENT APPLICATION THAT MET THE REQUIREMENTS TO BE GRANTED A FILING DATE.

APPLICATION NUMBER: 60/673,518

FILING DATE: *April 20, 2005*

RELATED PCT APPLICATION NUMBER: *PCT/US06/09103*

THE COUNTRY CODE AND NUMBER OF YOUR PRIORITY APPLICATION, TO BE USED FOR FILING ABROAD UNDER THE PARIS CONVENTION, IS *US60/673,518*



Certified by

Under Secretary of Commerce
for Intellectual Property
and Director of the United States
Patent and Trademark Office



042005

22764 U.S. PTO

Practitioner's Docket No. 100325.0284PRO

PATENT

Preliminary Classification
Proposed Class:
Subclass:

113277 U.S. PTO
60/673518

042005

IN THE UNITED STATES PATENT AND TRADEMARK OFFICE

In re application of: John Mak

For: Integrated NGL Recovery and LNG Liquefaction

Commissioner for Patents
P.O. Box 1450
Alexandria, VA 22313-1450

COVER SHEET FOR FILING PROVISIONAL APPLICATION
(37 C.F.R. § 1.51(c)(1))

This is a request for filing a PROVISIONAL APPLICATION FOR PATENT under 37 C.F.R. § 1.51(c)(1)(i). The following comprises the information required by 37 C.F.R. § 1.51(c)(1):

1. The following comprises the information required by 37 C.F.R. § 1.51(c)(1):
2. The name and residence of the inventor is (37 C.F.R. § 1.51(c)(1)(ii)(iii)):
 1. John Mak
2242 Salt Air Drive
Santa Ana, CA 92705
USA

EXPRESS MAILING UNDER 37 C.F.R. § 1.10*

(Express Mail label number is mandatory.)

(Express Mail certification is optional)

I hereby certify that this paper, along with any document referred to, is being deposited with the United States Postal Service on this date April 20, 2005 in an envelope addressed to the Commissioner for Patents, P.O. Box 1450, Alexandria, VA 22313-1450 as "Express Mail Post Office to Addressee" Mailing Label No. EV724398277US.

Date:

4/20/05

Sara L. Geer

Type or print name of person mailing paper

Signature of person certifying

WARNING: Certificate of mailing (first class) or facsimile transmission procedures of 37 C.F.R. 1.8 cannot be used to obtain a date of mailing or transmission for this correspondence.

***WARNING:** Each paper or fee filed by "Express Mail" must have the number of the "Express Mail" mailing label placed thereon prior to mailing. 37 C.F.R. 1.10(b).

"Since the filing of correspondence under § 1.10 without the Express Mail mailing label thereon is an oversight that can be avoided by the exercise of reasonable care, requests for waiver of this requirement will not be granted on petition." Notice of Oct. 24, 1996, 60 Fed. Reg. 56,439, at 56,442.

3. The title of the invention is (37 C.F.R. § 1.51(c)(1)(iv)):

Integrated NGL Recovery and LNG Liquefaction

4. The name, registration, customer and telephone numbers of the practitioner are (37 C.F.R. § 1.51(c)(1)(v)):

Name of practitioner: Martin Fessenmaier

Reg. No. 46697

Tel. 714-641-5100

Customer No. 34284

5. The docket number used to identify this application is (37 C.F.R. § 1.51(c)(1)(vi)):

Docket No. 100325.0284PRO

6. The correspondence address for this application is (37 C.F.R. § 1.51(c)(1)(vii)):

Martin Fessenmaier
611 Anton Blvd., Suite 1400
Costa Mesa, CA 92626
USA

7. Statement as to whether invention was made by an agency of the U.S. Government or under contract with an agency of the U.S. Government. (37 C.F.R. § 1.51(c)(1)(viii)).

This invention was NOT made by an agency of the United States Government, or under contract with an agency of the United States Government.

8. Identification of documents accompanying this cover sheet:

A. Documents required by 37 C.F.R. § 1.51(c)(2)-(3):

Specification:	No. of pages	14
Drawings:	No. of sheets	5

9. Fee

The filing fee for this provisional application, as set in 37 C.F.R. § 1.16(k), is \$200.00 for other than a small entity.

10. Fee payment

Fee payment in the amount of \$200.00 is being made at this time.

11. Method of fee payment

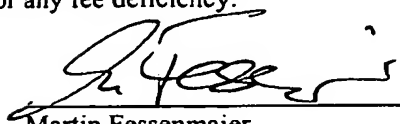
Charge Account No. 502191, in the amount of \$200.00.

A duplicate of this Cover Sheet is attached.

Please charge Account No. 502191 for any fee deficiency.

Date:

04/20/05



Martin Fessenmaier
Registration No. 46697
Rutan & Tucker, LLP
611 Anton Blvd., Suite 1400
Costa Mesa, CA 92626
USA
714-641-5100
Customer No. 34284

INTEGRATED NGL RECOVERY AND LNG LIQUEFACTION

Field of The Invention

The field of the invention is natural gas liquids (NGL) recovery and LNG liquefaction, and particularly integrated plant configurations for same.

5 Background of The Invention

While the crude oil supply in the world is diminishing, the supply of natural gas is still relatively abundant in many parts of the world. Natural gas is typically recovered from oil and gas production wells located onshore and offshore, and predominantly comprises C1 (methane). Depending on the particular formations and reservoirs, natural gas also contains relatively low
10 quantities of non-methane hydrocarbons, including C2 (ethane), C3 (propane), C4 (butane), C5 (pentane), and heavier components. Further components of natural gas include water, nitrogen, carbon dioxide, hydrogen sulfide, mercaptans, and other gases.

Natural gas from the wellheads is commonly treated and processed and transported to gas processing plants in high pressure transmission pipelines. However, in remote locations without the
15 necessary pipeline infrastructure, natural gas is commonly transported by liquefying the natural gas and transporting the gas in liquid form (*e.g.*, using LNG cargo carriers). Unfortunately, liquefaction of natural gas is problematic as natural gas also contains C5 and aromatics heavier hydrocarbons, which solidify when cooled to cryogenic temperatures. Consequently, most aromatic hydrocarbons must be removed to a low level (typically less than 1 ppmv) to avoid solidification and ultimately
20 plugging the cryogenic heat exchange equipment. Additionally, lighter hydrocarbons such as C2, C3 and C4 must also be removed at least to some degree for the North America market, which typically requires the heating value of the natural gas to meet a heating value (Higher Heating Value) between 1050 to 1070 Btu/SCF. There are also economic incentives to extract these hydrocarbons as they can be sold at a premium price over natural gas. For example, C2 can often be used as a feedstock for
25 petrochemical manufacture, C3 and C4 can be sold as LPG fuels, and the C5+ hydrocarbons can be used for gasoline blending.

There are numerous configurations and methods known in the art for C2 and C3+ NGL recovery from a natural gas feed. However, all past effort has been focused on removal of the NGL hydrocarbons from natural gas using standalone NGL recovery plants, which operate independently from LNG liquefaction plants. These processes generally produce a relatively low pressure residue gas at ambient temperature, which would necessitate re-compression and re-cooling of the residue gas in the LNG liquefaction plant. Typical examples include the expander processes described in U.S. Pat. Nos. 4,157,904 to Campbell et al., 4,251,249 to Gulsby, 4,617,039 to Buck, 4,690,702 to Paradowski et al., 5,275,005 to Campbell et al., 5,799,507 to Wilkinson et al., and 5,890,378 to Rambo et al.

For other high C2 recovery, some configurations as described in U.S. Pat. Nos. 6,116,050, require letting down a portion of the residue gas compressor discharge to the NGL recovery column as a methane rich reflux using the Joule-Thomson (JT) valve. While these processes can improve C2 recovery to some extent, additional residue gas compressor horsepower is required, which may render the process costly to operate. There are also more recent advances in the C2 and C3 recovery area (see *e.g.*, commonly owned Patent No. 6,837,070) in which a high pressure absorber is coupled with a lower pressure distillation column to improve NGL recovery efficiency. However, these NGL processes are designed for either high C2 or C3 recovery, and generally not designed for varying levels of C2 recovery without lowering C3 recovery. Thus, in most cases, standalone NGL recovery plants are generally used to produce a low pressure and ambient temperature residue gas that require re-compression and re-cooling in the LNG liquefaction plant, duplicating many of the refrigeration and heat exchange equipment in the NGL recovery plant.

In other known approaches, attempts were made to include the NGL recovery process as part of the LNG liquefaction plant, as disclosed in the U.S. Pat. Nos. 6,401,486 to Lee et al. and 6,662,589 to Roberts et al. Lee et al. teaches a methane rich stream can be used as reflux to the NGL recovery column coupled with an overhead condenser in the another NGL column to achieve an propane recovery of 95%. However this configuration requires the NGL column operating at a pressure of 450 psig or even lower, as the separation of NGL becomes increasingly difficult at the higher pressures due to the correspondingly reduced relative volatility. Consequently, these

processes require significant recompression from the column overhead to the required LNG liquefaction pressure, typically from 450 psig to about 800 psig to 900 psig.

Roberts et al in Pat. No. 6,662,589 teach a C2 rich liquid being recycled from the NGL fractionation unit that is used for C3 absorption in a high pressure absorption column. While this process attempts to operate the NGL column at a high pressure (e.g., 600 psig), NGL separation efficiency suffers as the relative volatility of the NGL components is reduced, which results in recovering significantly less NGL components, especially C2 components. Without removal of a high level of C2 and C3 components, the currently known processes cannot produce a lean natural gas with sufficiently lower heating value content to meet the North America pipeline specifications in an economically manner. Additionally, the lean gas pressure from these processes would require significant refrigeration in the LNG liquefaction plant due to the relatively low feed gas pressure. LNG liquefaction generally requires significantly less refrigeration duty when operating at a higher pressure, between 800 psig and 900 psig or higher.

Presently known LNG liquefaction processes generally include variations of steps in which the natural gas is cooled and condensed, using either pure components or mixed refrigerants in different configurations. One of the variations is the cascade refrigeration process that employs heat exchange of the natural gas with several pure component refrigerants having successively lower boiling points, such as propane, ethane, and methane, or alternatively with a mixed refrigerant. The heat exchange process can also be accomplished using a single refrigerant mixture comprising distinct components by evaporating the refrigerant at several different pressure levels. Alternatively, natural gas cooling can also be achieved by expansion of the natural gas using either Joule-Thomson expansion or an expansion turbine.

Thus, while numerous compositions and methods for NGL recovery and LNG liquefaction are known in the art, all or almost all of them, suffer from one or more disadvantages. Therefore, there is still a need for improved NGL recovery and LNG liquefaction.

Summary of the Invention

In accordance with the present invention, it has been found that a high pressure cryogenic vapor stream predominantly comprising methane can be produced from a NGL recovery plant that advantageously reduces, or even eliminates the duplication of refrigeration and heat exchange steps and equipment in the LNG liquefaction plant. Such configurations and methods greatly reduce the refrigeration requirement in the LNG liquefaction plant, while advantageously allowing recovery of 99% propane and of up to 85% ethane from the feed gas. Contemplated NGL recovery plants will produce a lean gas predominantly comprising methane with a heating value that meets the North America natural gas pipeline requirement, and consequently, suitable for integrating to an LNG liquefaction plant, which will in turn increase the throughput of the LNG liquefaction train. Among other advantages, contemplated configurations can be changed from C3 recovery to C2 recovery by adjusting a flow split ratio between the top reflux and the expansion flow while diverting at least a portion of the absorber bottoms product flow to a distillation column.

Viewed from a different perspective, methods and configurations of an NGL recovery plant allow production of a high pressure cryogenic methane rich vapor for feeding an LNG liquefaction plant, while recovering 99% of the propane and up to 85% ethane from the feed gas. In such plants, the gas processing portion comprises a refluxed absorber producing a bottom stream and receiving a feed gas and an absorber reflux stream that is produced from the overhead vapor from a distillation column (preferably after the overhead vapor is compressed and cooled). In preferred configurations, a distillation column is fluidly coupled to an absorber, receives a column feed stream and operates at a pressure that is at least 50 to 100 psi lower, more preferably 100 psi to 300 psi lower than the operating pressure of the absorber. Seamless changeover of the C3 recovery operation to the C2 recovery operation (or vice versa) while maintaining 98% or higher C3 recovery for any level of C2 recovery is achieved by increasing the second reflux to the absorber while reducing the flow to the expander, with simultaneously diverting at least a portion of the absorber bottoms product flow to the distillation column.

Consequently, with two columns operating at different pressures, a very high recovery of the C2 plus components is possible and the high pressure column operation is beneficial in reducing the

refrigeration duty in the LNG liquefaction plant. **Figure 5** illustrates some of the advantages of the contemplated configurations and methods in which relative volatilities of the NGL components (that is C1 to C2) in the distillation column (s) are compared to known configurations. Here, both the present configurations and the known configurations are compared for a column operating at 600 psig pressure for C2 recovery. The relative volatilities of the NGL components of the known configurations are represented by curve A, which drops to a very low value of 2 in the mid section of the column. These low relative volatilities are the primary reason of lower separation efficiency and lower NGL recoveries, even with a large number of fractionation trays. In contrast, employing two columns operating at different pressures, with the first column (absorber) operating at a high pressure of 600 psig (curve B) and the second column operating at 450 psig (curve C), dramatically increases relative volatilities of the NGL components (*e.g.*, to values of over 10), resulting in higher separation efficiency and higher NGL recoveries.

Therefore, in one aspect of the inventive subject matter, the distillation column comprises a demethanizer or deethanizer column, the feed gas has a pressure of between 900 psig and 1600 psig, and is expanded in a turbo-expander to an absorber. The bottoms product of the absorber is expanded in a range of 50 psi to 350 psi, thereby chilled by Joule-Thomson effect to -90 °F to -130 °F. It is also contemplated that the cooled and expanded bottoms product stream is fed as the distillation column feed stream into the distillation column, and it is further contemplated that the expanded bottoms product stream may further provide cooling for the feed gas and reflux to the column, and at least of a portion of the expanded absorber bottoms product may also be routed directly to the distillation column for C2 absorption during the C2 recovery operation.

In a particularly contemplated aspect, the distillation column produces an overhead stream, which is compressed, cooled, and fed to the absorber as the first absorber reflux stream. It is also contemplated that the compressor can be located in various locations between the absorber and distillation column and can be fed to various locations in the absorber for C3 recovery and/or C2 recovery. In a further contemplated aspect, an external refrigeration unit is fluidly coupled to feed gas exchangers and reflux exchangers to supply feed gas chilling and column reflux duties. The external refrigeration unit can be accomplished by any types of refrigeration systems, with pure

components or mixed refrigerants using different methods. For example, a cascade refrigeration process may employ heat exchange of the natural gas with several pure component refrigerants having successively lower boiling points, or a heat exchange process may be accomplished using a single refrigerant with multiple pure components by evaporating the refrigerant at several different pressure levels. Alternatively, or additionally, natural gas cooling can also be achieved by expansion of the natural gas using either Joule-Thomson expansion or expansion turbine.

In especially contemplated aspects, the absorber is configured to separately receive a first and a second portion of a feed gas vapor, and a distillation column overhead, wherein the first portion of the feed gas vapor and the distillation column overhead provide reflux to the absorber. In such configurations, a flow control adjusts the ratio of at least one of the first and second portions of the feed gas vapor to produce the desired recovery levels of ethane. Among other advantages, it should be recognized that an optimum flow split ratio of the first and second flow of the feed gas is employed for the various C2 recovery while maintaining a high C3 (98% or above) recovery.

It is still further contemplated that at least a portion of the absorber bottoms product is used as a lean oil for C2 absorption in the distillation column during C2 recovery, and in this preferred configuration, flow to the feed cooler is reduced or stopped, directing the absorber bottoms to the distillation column. C2 recovery in such configurations increases when the first portion of the feed gas vapor increases relative to the second portion of the feed gas vapor. Thus, preferred configurations permit a seamless changeover of the C3 recovery operation to the C2 recovery operation (or vice versa) while maintaining 98% or higher C3 recovery.

Consequently, in another aspect of the inventive subject matter, a method of operating a plant includes a step of providing an absorber and a distillation column, wherein the absorber receives a plurality of absorber feed streams and provides a bottom product to the distillation column. In another step, at least one of the feed streams is split into a first and second portion, wherein the first and second portions are introduced into the absorber at different locations, and in still another step, the flow ratio between the first and second portions is used to control the degree of recovery of the C2 component in the bottom product of the distillation column. In yet a further step,

the absorber bottoms product is used as a lean oil for absorption of the C2 component in the distillation column.

In one aspect of the inventive subject matter, the absorber further receives a liquid portion of the natural gas feed and a second vapor portion of the natural gas feed, wherein the second portion is reduced in pressure via a turbo expander. Preferred absorbers produce a bottom product that is letdown and cools at least one of the feed gas and the reflux streams, and all or at least a portion of the bottom product may be fed into the distillation column. Any types of refrigeration cycles can be employed in the cooling process, with a mixed refrigeration system or pure component refrigeration cascade system.

In a still further contemplated aspect, the absorber produces an overhead vapor product that is predominantly methane at cryogenic temperature (-80°F or lower), which is further compressed using power generated by the turbo-expansion of the feed gas. Such configuration produces a high pressure cryogenic vapor at 800 psig to 900 psig or higher that can advantageously be directly fed to the LNG liquefaction plant. It should be recognized that compressing a cryogenic vapor from the high pressure absorber is relatively energy efficient, and therefore allows operation of an LNG liquefaction plant with relatively low, and more typically no external compression of the feed gas.

It is also contemplated that an external refrigeration unit can be employed to cool at least one of the first and second reflux streams, and may further cool at least one of the natural gas feed and a vapor portion of the natural gas feed. Preferred pressure letdown devices include turbo-expander and the Joule-Thomson valves, and preferred distillation column comprises a demethanizer or deethanizer. Where C2 recovery is particularly preferred, it is contemplated that the first lean reflux stream from the feed gas may be fed into the absorber as a subcooled liquid (a liquid that is cooled below its bubble point temperature), wherein the distillation column comprises a demethanizer.

Consequently, in a further aspect of the inventive subject matter, a method of increasing throughput in the LNG liquefaction plant using a natural gas recovery plant having an absorber and a distillation column includes one step in which a first reflux stream is provided to the absorber,

wherein the first reflux stream comprises an overhead product from the distillation column. In another step, a bypass is provided upstream of a turbo expander, wherein the bypass receives a vapor portion of a cooled natural gas and provides the vapor portion to the absorber, and in yet another step, the pressure of the vapor portion is cooled, at least partially condensed and reduced in pressure before the vapor portion enters the absorber as a second reflux stream. In a still further step, a heat exchanger is provided that cools at least one of the first and second reflux streams using an external refrigerant.

Therefore, a method of operating a plant may include one step in which an absorber and a distillation column are provided. In a further step, a cooled lean overhead product from the distillation column is fed to the absorber as a first reflux stream, and in another step, the pressure of a cooled vapor portion of a natural gas feed is reduced via a device other than a turbo expander, wherein the cooled vapor portion that is at least partially condensed or subcooled, reduced in pressure and fed to the absorber as a second reflux stream.

Various objects, features, aspects and advantages of the present invention will become more apparent from the accompanying drawing and the following detailed description of preferred embodiments of the invention.

Brief Description of the Drawing

Figure 1 is a Prior Art Figure depicting a schematic of an exemplary plant configuration.

Figure 2 is a schematic of an exemplary plant configuration according to the inventive subject matter.

Figure 3 is a graph depicting composite heat curves for heat exchanger 51 and 54 in a plant according to Figure 2 for C3 recovery.

Figure 4 is a graph depicting composite heat curves for heat exchanger 51 and 54 in a plant according to Figure 2 for C2 recovery.

Figure 5 is a graph comparing the relative volatilities of the NGL components at 600 psig between the prior art and a plant according to Figure 2 in C2 recovery.

Table 1 is a table depicting the overall mass balance for C3 recovery when the plant is operated on C3 recovery mode.

5 Table 2 is a table depicting the overall mass balance for C2 recovery when the plant is operated on C2 recovery mode.

Detailed Description

The inventor discovered an efficient and flexible configuration and process that produces a high pressure cryogenic vapor stream (containing predominately methane) suitable for feeding to an LNG liquefaction plant, and that further produces a liquid stream containing predominantly ethane and heavier hydrocarbons. Contemplated configurations and processes can achieve 99% propane recovery when operating in a propane recovery mode, and can also achieve up to 85% ethane recovery when operating in an ethane recovery mode without substantial reduction (less than 3% absolute in reduction) in propane recovery. Contemplated configurations and processes can also provide a seamless and gradual changeover method of the C3 recovery operation to the C2 recovery operation (or *vice versa*) by adjusting only the flow streams to the absorber and the distillation column while such changeover configuration results in 98% or higher C3 recovery in all C2 recovery operation.

Viewed from a different perspective, the inventor discovered that high propane recovery (i.e., at least 95%) and high ethane recovery (up to 85%) from a feed gas with relatively high pressure (e.g., between about 800 psig to 1600 psig) can be realized by operating an absorber in a gas processing plant at a higher pressure than a distillation column (e.g., a demethanizer or deethanizer), and in which a compressor is used to recycle the distillation column overhead to the absorber. The absorber bottoms product is preferably expanded to provide cooling for the feed gas and the reflux stream. The overhead vapor from the absorber is then further compressed forming a feed gas to an LNG liquefaction plant. Further configurations related to some aspects of the

inventive subject matter are disclosed in our copending U.S. patent application with the serial number 10/478705, which is incorporated by reference herein.

Prior Art Figure 1 shows a standalone C2 NGL recovery process that is coupled with a standalone LNG liquefaction plant. Here, contaminants free and dried feed gas stream 1, typically supplied at about 1200 psig, is cooled in exchanger 51, using column overhead vapor, side reboiler stream 22 and external refrigerant 32. Liquid is removed from separator 52 and is sent to the NGL column 58 that acts as a demethanizer. The flashed vapor from separator 52 is split into two portions, one portion is cooled in the exchanger 54 to provide reflux to the column, and the other portion is expanded, cooled and sent to the lower section for rectification. Consequently, it should be noted that the above standalone gas subcooled process produces a residue gas at ambient temperature and about 450 psig which must be recompressed using re-compressor 100 prior to the standalone LNG plant. It should further be especially appreciated that the above non-integrated NGL recovery plant and LNG liquefaction plant is relatively inefficient as additional power is required by the residue gas compressor, and as the feed gas cooler 51 must be duplicated in the LNG liquefaction plant.

In contrast, a preferred configuration of a gas processing plant for integrated NGL recovery plant and LNG liquefaction is depicted in **Figure 2**, which can be used for flexible C3 and C2 recovery. Exemplary compositions, temperatures and pressures, and flow rates of feed gas, product gas, and liquid product for exemplary and typical operation are shown in **Table 1** and **Table 2** for C3 and C2 recovery, respectively.

In **Figure 2**, feed gas stream 1 enters the plant at about 1200 psig and 120 °F, and is cooled in heat exchanger 51 to typically -10 °F to -40 °F, forming stream 2, using multiple cooling streams, from the letdown absorber bottoms stream 15, liquid stream 5 from separator 52, side reboiler stream 22 from the distillation column and an external refrigeration stream 32. Any type of available refrigeration systems is applicable, including the pure component cascade refrigeration cycles or the mixed refrigerants cycles or a combination of both system. The NGL process also generates refrigeration with expansion using turbo-expander and the Joule-Thomson (JT) valves. The high

energy efficiency of this process can be demonstrated by the close temperature approaches of the heating and cooling curves in the minimization of loss work. The combined hot composite curve and the combined cold composite curve of the feed gas exchanger 51 and the reflux exchanger 54 are shown in **Figure 3** and **Figure 4** for C3 and C2 recovery, respectively.

5 The chilled feed gas stream 2 is separated in a separator 52, forming a gaseous portion 3 and a liquid portion 4. The liquid portion 4 is letdown in pressure in JT valve 53 forming stream 5 typically at about -40 °F. During C3 recovery, stream 5 is heated in exchanger 51 to about 80 °F forming stream 6, using the heat content in the feed gas. Stream 6 enters the stripping section of the distillation column 61 for removal of the C2 and lighter components. The gaseous portion 3 from
10 separator 52 is split into two portions. One portion is routed to the exchanger 54 to provide reflux to the absorber, and the other portion is expanded in turbo-expander 64 which generates power to operate compressor 65 while producing a chilled vapor stream 10, typically at -80 °F to -100 °F. The chilled vapor is letdown in pressure to the absorber 58, which operates at 500 psig to 700 psig, typically at 600 psig.

15 For the different C2 recovery levels, the flow split ratio, that is, stream 8 to stream 3, can be adjusted to maintain a high C3 recovery. The following table shows the different split ratios and the results of C3 and C2 recovery; and for all C2 recovery levels, high C3 recoveries (over 98%) are maintained.

SPLIT RATIO (RATIO OF STREAM 8 TO STREAM 3)	C3 RECOVERY, %	C2 RECOVERY, %
0.7	98	85
0.8	98	62
0.9	99	31
1.0	99	25

20 Absorber 58 is refluxed with two cold streams, with a first reflux (top reflux) supplied by stream 27 from the distillation column 61, and a second reflux stream 12 from exchanger 54. With this twin reflux streams and an optimum split flow ratio, high C3 recovery can be maintained for the various levels of ethane recovery. During the C3 recovery, the C2 content in the NGL product from

the distillation column is lowered by increasing the bottom temperature using heat supplied from the side reboiler and the bottom reboiler 63.

The absorber produces an overhead vapor stream 28 at about -100 °F to -110 °F and a bottoms stream 14 at about -90 °F to -100 °F. The overhead vapor is compressed by the residue gas compressor 65 using power generated by turbo-expander 64 forming a discharge stream 29, typically at about 900 psig and -70 °F to -80°F. It should be especially appreciated that compression of a cold vapor is energy efficient as the achievable compression ratio across the compressor is significantly higher than that using a warm vapor of the standalone plant. Subsequently, the process produces a high pressure and cryogenic temperature vapor stream that can be fed to the LNG liquefaction plant 67 for LNG production forming stream 30 at about -260 °F.

The absorber bottoms stream 14 is letdown in pressure in JT valve 59 to about 460 psig, and is chilled to about -100 °F forming stream 15. During C3 recovery, this cold stream is used to provide a least a portion of the cooling duty of feed exchanger 51 and the reflux duty in condenser 62. The overhead condenser 62 consists of a heat exchange coil that is integral to the distillation column, generating an internal reflux stream 19 to the rectification section of the distillation column. Alternatively, the integral condenser 62 system can be replaced by an external system which would include an external heat exchanger, a separator and a reflux pump.

During C2 recovery, the absorber bottom stream 14 is routed directly to the top of the distillation column for absorption of the C2 and heavier components in the distillation column. In this operation, JT valve 59 is closed and JT valve 60 is open, routing the cold absorber bottoms to the distillation column for C2 recovery.

The temperature profile of distillation column 61 varies between C3 recovery and C2 recovery as exemplified in the following table.

TEMPERATURE PROFILE OF 61	C2 RECOVERY	C3 RECOVERY
Top (Stream 24)	-100 to -120°F	-35 to -45°F
Bottom (Stream 25)	-90 to -120°F	240 to 270°F

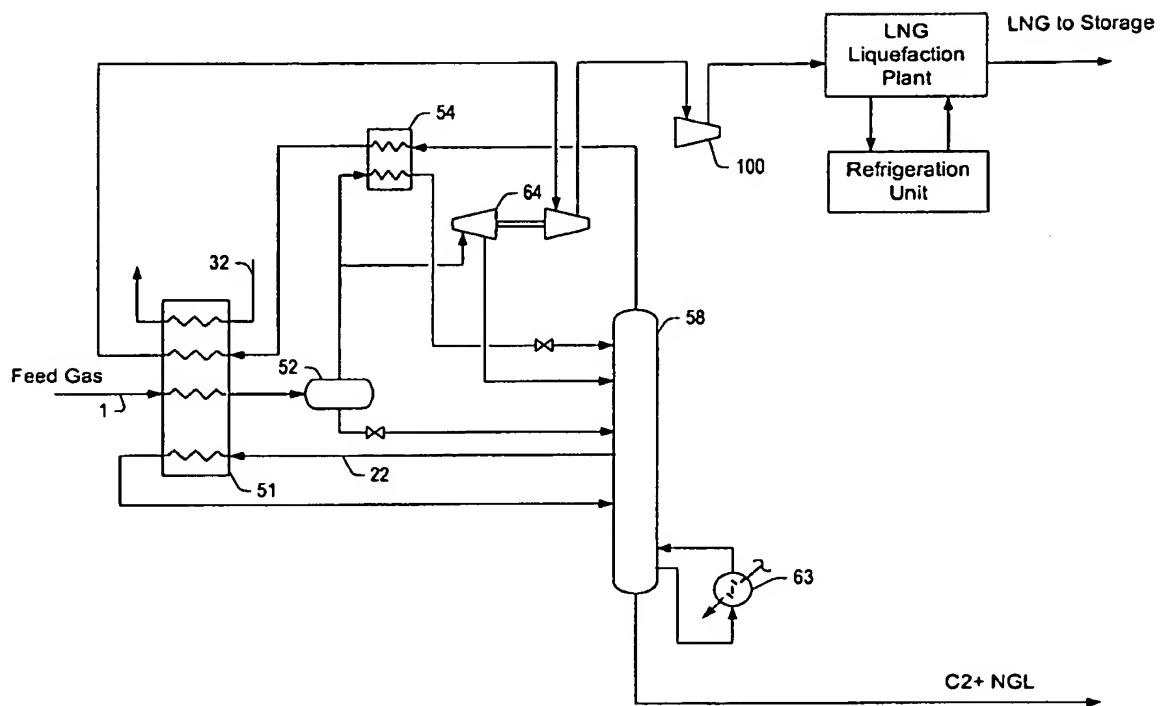
The NGL product specification is controlled with side reboiler integral to exchanger 51 and a bottom reboiler 63 using an external heat source 34. The distillation column produces an NGL bottoms products (C2 plus or C3 plus) and an overhead vapor stream 24 that is compressed in compressor 66 to about 600 psig, or as needed to enter the absorber as a top reflux, further cooled in heat exchanger 54, and then used as the first (top) reflux 27 to the absorber. Refrigerant stream 31 is supplied from an external refrigeration unit to the exchanger 54 for cooling and partially or totally condensing this recycle stream. It should be appreciated that recycling the overhead stream from the distillation permits recovery of the desirable NGL components allowing the distillation column to operate at the most efficient pressure for fractionation ~~of~~or separation of the desirable NGL components, as previously demonstrated in the hot and cold composite curves in Figure 3 and Figure 4.

With respect to the feed gas streams it is contemplated that numerous natural and man made feed gas streams are suitable for use in conjunction with the teachings presented herein, and especially preferred feed gas streams include natural gases, refinery gases, and synthetic gas streams from hydrocarbon materials such as naphtha, coal, oil, lignite, etc. Consequently, the pressure of contemplated feed gas streams may vary considerably. However, it is generally preferred that appropriate feed gas pressures for plant configurations according to Figure 2 will generally be in the range between 800 psig and 1600 psig, and that at least a portion of the feed gas is expanded in a turboexpander to provide cooling and/or power for the residue gas recompression.

With respect to all components of contemplated configurations (e.g., heat exchangers, pumps, valves, compressors, expanders, refluxed absorbers, de-ethanizers, etc.) it should be appreciated that all known and commercially available components are suitable for use in conjunction with the teachings presented herein. It is further generally contemplated that configurations according to the inventive subject matter may find wide applicability in gas plant

applications where high propane and ethane recovery are desirable, and feed gas is available at pressure greater than 800 psig. Moreover, such configurations produce a high pressure cryogenic methane rich vapor that will advantageously reduce equipment and operating costs when integrated to a LNG liquefaction plant.

5 Thus, specific embodiments and applications for high propane and ethane recovery processes and configurations have been disclosed. It should be apparent, however, to those skilled in the art that many more modifications besides those already described are possible without departing from the inventive concepts herein. The inventive subject matter, therefore, is not to be restricted except in the spirit of the appended claims. Moreover, in interpreting both the specification and the claims,
10 all terms should be interpreted in the broadest possible manner consistent with the context. In particular, the terms “comprises” and “comprising” should be interpreted as referring to elements, components, or steps in a non-exclusive manner, indicating that the referenced elements, components, or steps may be present, or utilized, or combined with other elements, components, or steps that are not expressly referenced.



Prior Art Figure 1

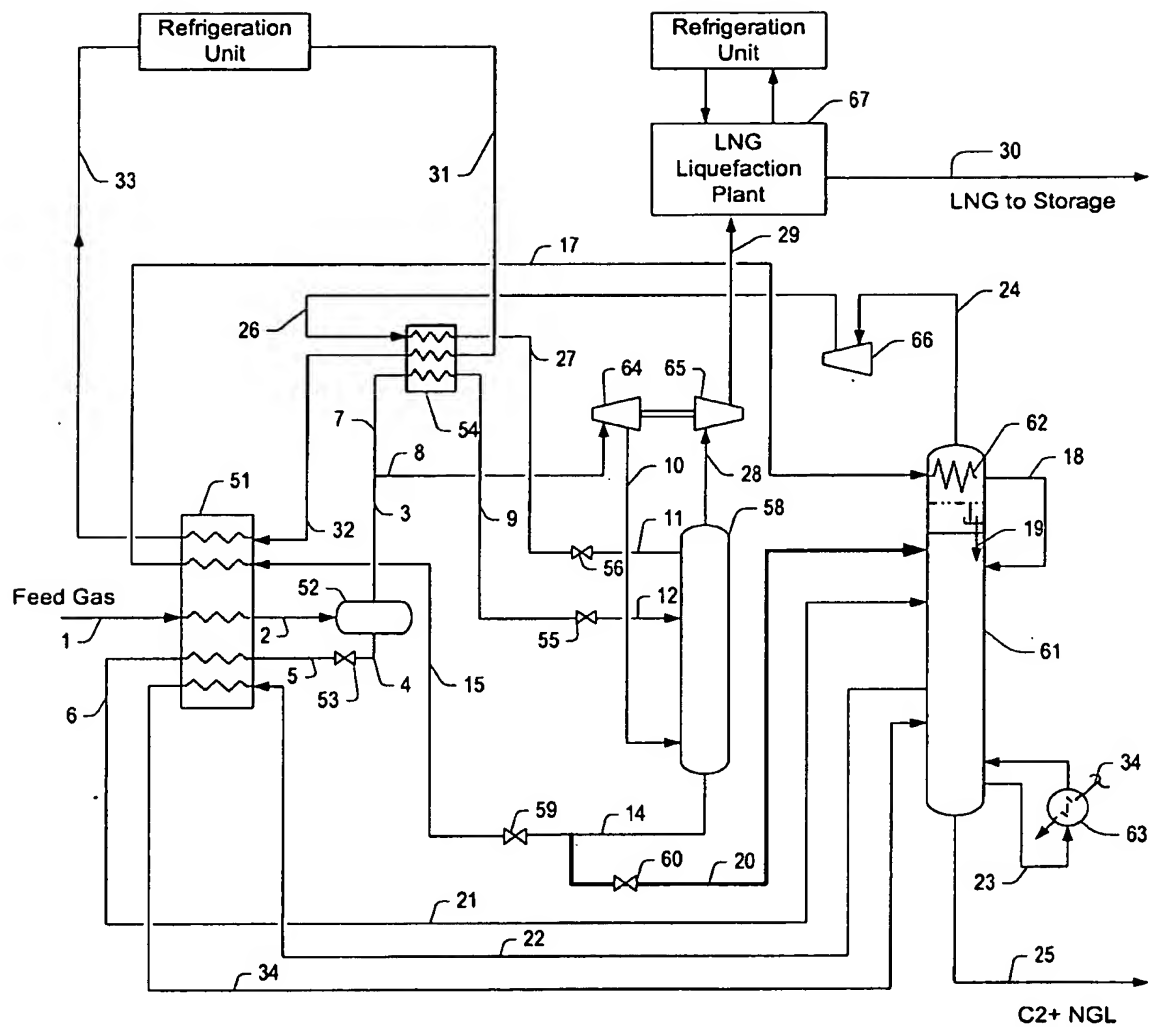


Figure 2

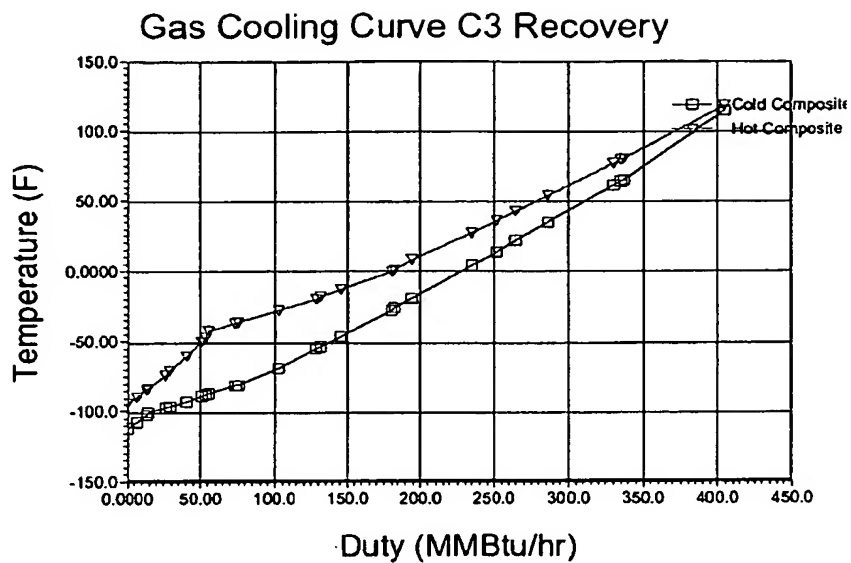


Figure 3 – C3 Recovery

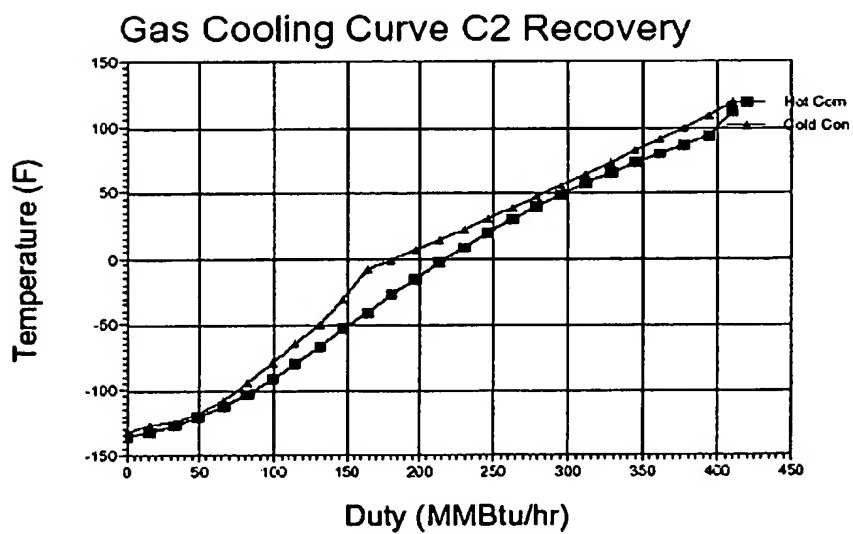


Figure 4 - C2 Recovery

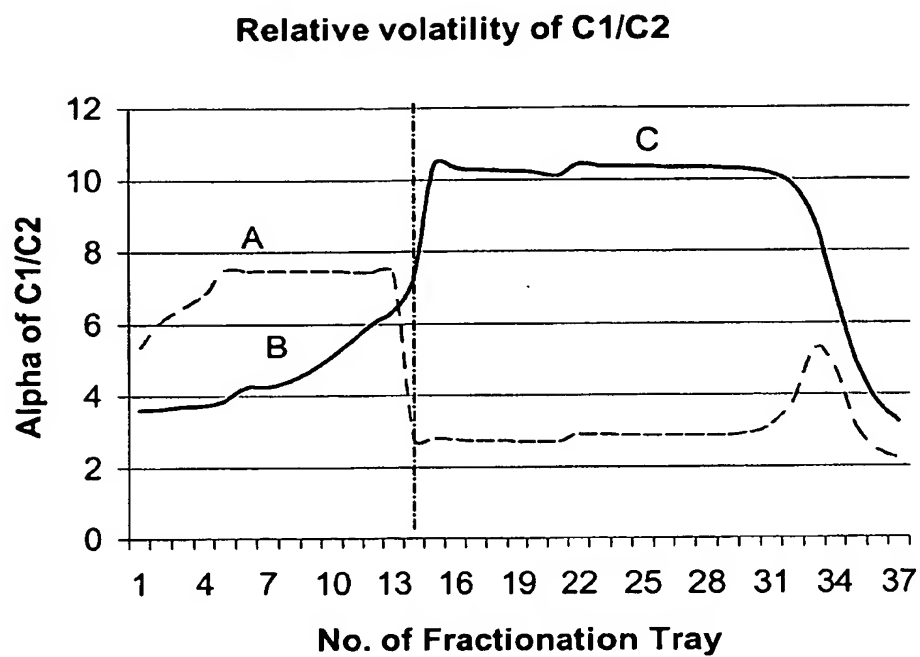


Figure 5

Stream	Feed	Liquid from NGL Recovery	Residue Gas to Liquefaction
CO2	0.000	0.000	0.000
N2	4.569	0.000	5.007
C1	86.161	1.020	94.320
C2	5.046	51.017	0.641
C3	1.854	21.093	0.011
iC4	0.395	4.514	0.000
nC4	0.590	6.751	0.000
iC5	0.248	2.833	0.000
Nc5	0.205	2.342	0.000
C6	0.224	2.565	0.000
C7	0.662	7.570	0.000
MMscfd	1,227	107	1,119
BPD		75,743	
Temperature, F	120	115	-75
Pressure, psig	1,200	470	900

Table 1- C3 Recovery

Stream	Feed	Liquid from NGL Recovery	Residue Gas to Liquefaction
CO2	0.000	0.000	0.000
N2	4.569	0.000	4.768
C1	86.161	0.000	89.906
C2	5.046	0.867	5.228
C3	1.854	43.345	0.051
iC4	0.395	9.478	0.000
nC4	0.590	14.173	0.000
iC5	0.248	5.946	0.000
Nc5	0.205	4.917	0.000
C6	0.224	5.384	0.000
C7	0.662	15.889	0.000
MMscfd	1,227	51	1,175
BPD		40,371	
Temperature, F	120	258	-60
Pressure, psig	1,200	440	886

Table 2 – C2 Recovery